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Control Benchmark for Solvent Recovery by Distillation

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Introduction

The need for tight control of a process has always been of great importance, especially within the field of distillation control, where great savings in energy and operational cost often can be obtained by using a suitable control strategy. However the transfer of research results in control design to the process industry is sometimes hindered by the complexity of the new developments.

The purpose of this contribution is to compare the practical consequences of three different control strategies on a real case-study, considering the ability of the resulting controllers to achieve the operational goals, as well as the complexity associated with this implementation.

The investigated strategies have been chosen to cover a broad range of control strategies and levels in the control hierarchy. The chosen controllers are:

- A classical decentralised two-input, two-output (TITO) PI control, aimed at the regulation of the system
- A TITO self-optimising control [1] with regulatory and economic evaluation
- Offset free model predictive allowing centralized regulation of the process [2]

Metods

The three control strategies are tested on a distillation model of an actual industrial process unit consisting of a distillation column, recovering solvent grade ethanol from an industrial waste stream containing water, ethanol (EtOH) and trace amounts of acetaldehyde (AA). The column is in operation continuously but the feed undergoes large step changes since the production upstream is batch based. The model is developed in MATLAB.

The pairing of the controller in the PID strategy is performed by using the Relative Gain Array (RGA) method [3]. The tuning of the PID strategy is obtained using the Internal Model Control (IMC) method by Garcia et al. [4] and Rivera et al. [5]. The self-optimizing controller (SOC) is designed following the null space [6]. For tuning and implementing the SOC strategy the same IMC method is used as with the PID control strategy.

The implementation of the MPC is based on an algorithm given by Huusom et al. [7], and the offset free MPC equations used are given by Rawlings and Mayne [2]. The tuning is done by trial and error since no tuning scheme seems to have been established to implement the offset free MPC method used [8].

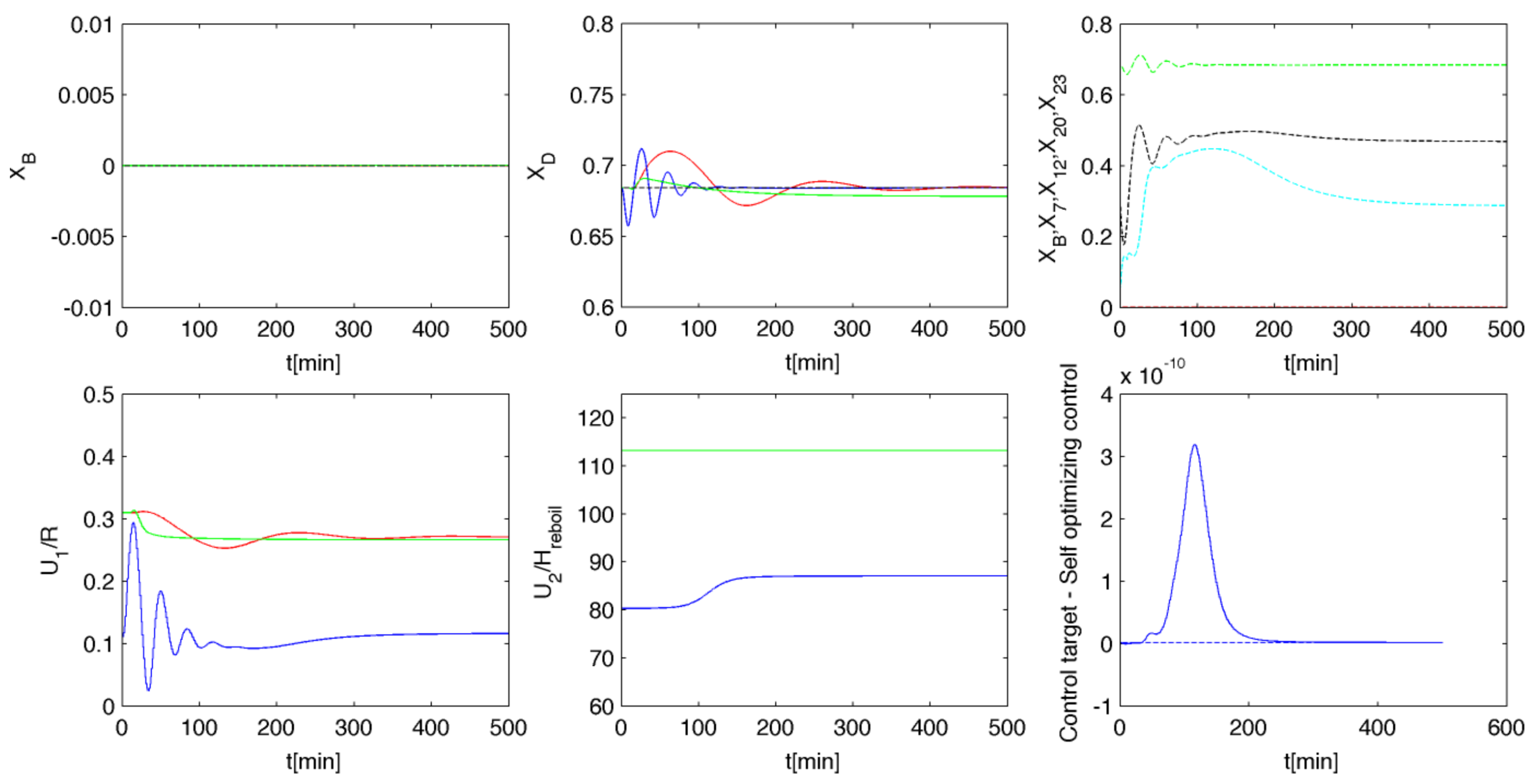

Figure 1: Disturbance of +15% in the amount of ethanol in the feed, (•) PID controller, (•) Self-optimized control, (•) Model Predictive Control, (−−) X_B, (−−) X_7, (−−) X_{12}, (−−) X_{20}, (−−) X_{23}

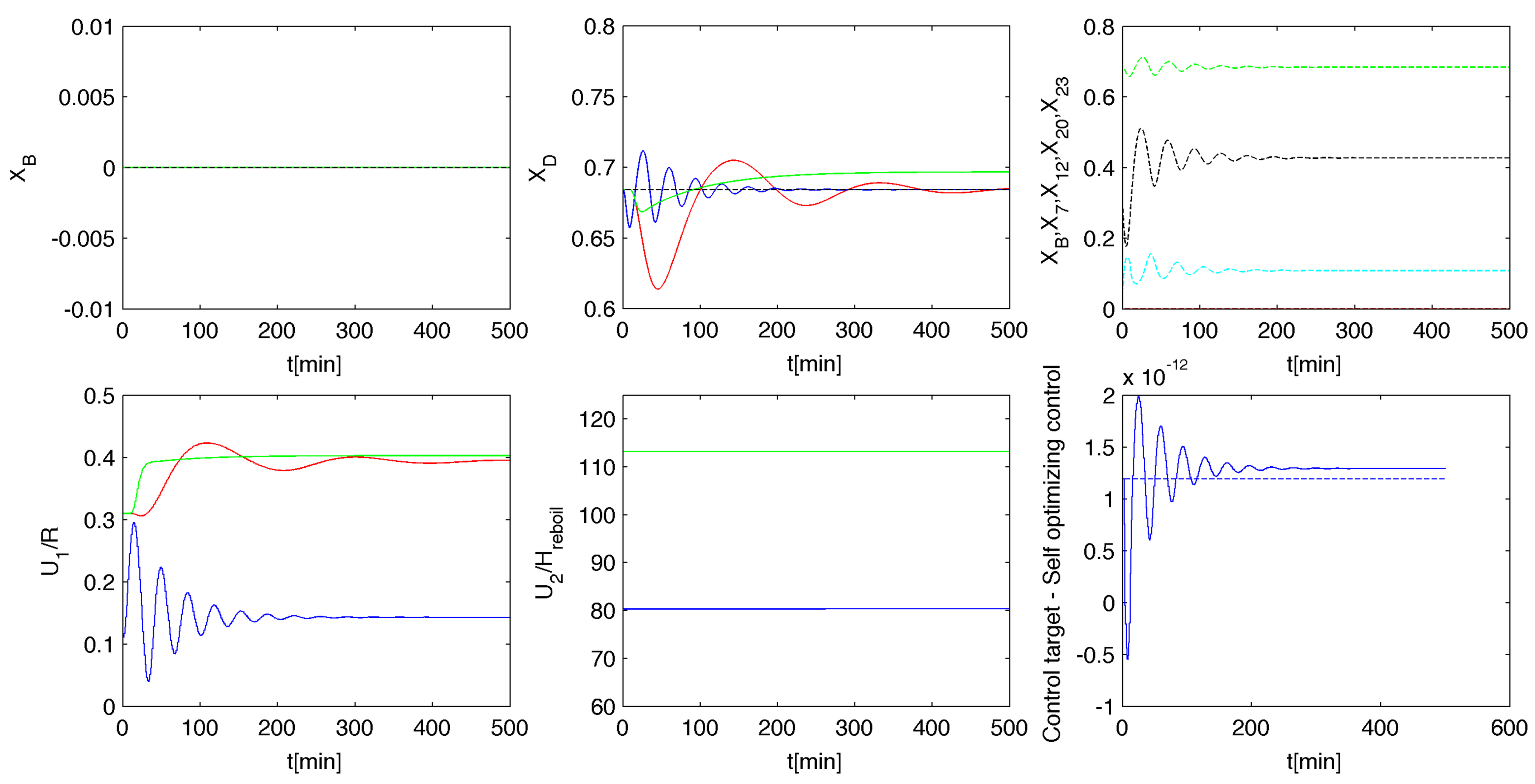

Figure 2: Disturbance of +7°C in the temperature feed, (•) PID controller, (•) Self-optimized control, (•) Model Predictive Control, (−−) X_B, (−−) X_7, (−−) X_{12}, (−−) X_{20}, (−−) X_{23}

	X_B - Integral of Absolute Error [mole frac.]			X_D - Integral of Absolute Error [mole frac.]		
	PI	SOC	MPC	PI	SOC	MPC
z_F +15%	9.52×10^{-11}	4.22×10^{-8}	9.57×10^{-11}	2.91	1.07	2.19
z_F -15%	5.84×10^{-11}	7.90×10^{-10}	5.82×10^{-11}	4.35	3.01	2.25
t_F +7°C	5.59×10^{-11}	6.72×10^{-10}	5.57×10^{-11}	5.83	1.40	4.75
t_F -7°C	1.02×10^{-10}	6.04×10^{-8}	1.03×10^{-10}	6.66	1.81	3.89
F +20%	2.81×10^{-10}	2.24×10^{-7}	2.89×10^{-10}	4.97	1.38	3.56
F -20%	3.34×10^{-11}	4.38×10^{-10}	3.30×10^{-11}	11.92	1.53	5.45

	U_1/Reflux – Total Variation [kmol]			U_2/Reboiler – Total Variation [kmol]		
	PI	SOC	MPC	PI	SOC	MPC
z_F +15%	0.10	0.83	0.05	3.68×10^{-6}	6,69	0.01
z_F -15%	0.15	2.34	0.05	2.26×10^{-6}	0,05	0.01
t_F +7C	0.20	1.09	0.09	2.16×10^{-6}	0,03	0.03
t_F -7C	0.23	1.42	0.08	3.94×10^{-6}	7,86	0.02
F +20%	0.16	1.07	0.18	1.09×10^{-5}	24,79	0.02
F -20%	0.39	1.18	0.41	1.29×10^{-6}	0,05	0.04

Table 1: Integral of absolute error and total variation over 400 minutes

	PI [currency]	SOC [currency]	MPC [currency]
z_F +15%	4406.9	5636.16	4510.5
z_F -15%	1912.9	3403.22	1830.1
t_F +7°C	3268.5	4715.37	3061.9
t_F -7°C	3087.3	4290.49	3246.4
F +20%	3905.7	4974.00	4199.8
F -20%	2484.3	3769.09	2022.9

Table 2: Integral of profit function over 400 minutes

Results and Disussion

The control systems behavior on a disturbance are shown in the figures.

The result of the 3 evaluation methods applied on the different disturbances is given in Table 1-2. As seen in the figures, all control strategies are able to stabilise the system, and bring the measured variables back or close to the desired set point, but the SOC need more aggressive use of the manipulated variables, compared to the other strategies.

The evaluation is performed at two levels: First the single controller is evaluated and later the entire system. Comparing the result of the IAE given in Table 3 for the two measured points it is clear that all the controllers are able to keep the concentration at the bottom of the column constant.

However, from comparing the IAEs at the top it is clear that the self-optimizing controller shows the largest IAE. However, this is to be expected since the controller finds the optimal set point based on a profit function, calculating the maximum profit, rather than a function, where the offset from the set-point is minimized.

The second parameter on which the strategies are compared is the total variation of the manipulated variables. All three control strategies have a very low TV when manipulating the reflux of the column. Looking at the TV of the energy given to the reboiler the self-optimizing control obtains much higher values, showing that this strategy has a large need to change the energy in order to obtain a stable and optimal control.

The last level is the overall comparison, where the integral of the profit is compared. As shown in Table 4 the self-optimizing controller is the control strategy that obtains the largest profit when a disturbance is imposed on the system.

Conclusion

From the data presented it is clear that the PID is the easiest to implement, but not the most accurate controller, neither in offset nor cost. The SOC is the best controller for minimizing the cost of a disturbance, but also the most aggressive controller. The MPC shows really good and accurate control, but is very heavy to implement and requires online optimization. The simplicity of implementing the self-optimizing control theory, the lack of online calculations, and the usage of a profit function make this the best suited control strategy for the distillation column investigated.

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